



Thermodynamic analysis of SOFC (solid oxide fuel cell)–Stirling hybrid plants using alternative fuels



Masoud Rokni

Technical University of Denmark, Department of Mechanical Engineering, Thermal Energy System, Copenhagen, Denmark

ARTICLE INFO

Article history:

Received 17 September 2012

Received in revised form

28 May 2013

Accepted 2 June 2013

Available online 1 July 2013

Keywords:

SOFC

Stirling

Fuel cell

Hybrid systems

System efficiency

Multi-fuel

ABSTRACT

A novel hybrid power system (~ 10 kW) for an average family home is proposed. The system investigated contains a solid oxide fuel cell (SOFC) on top of a Stirling engine. The off-gases produced in the SOFC cycle are fed to a bottoming Stirling engine, at which additional power is generated. Simulations of the proposed system were conducted using different fuels, which should facilitate the use of a variety of fuels depending on availability. Here, the results for natural gas (NG), ammonia, di-methyl ether (DME), methanol and ethanol are presented and analyzed. The system behavior is further investigated by comparing the effects of key factors, such as the utilization factor and the operating conditions under which these fuels are used. Moreover, the effect of using a methanator on the plant efficiency is also studied. The combined system improves the overall electrical efficiency relative to that of a stand-alone Stirling engine or SOFC plant. For the combined SOFC and Stirling configuration, the overall power production was increased by approximately 10% compared to that of a stand-alone SOFC plant. System efficiencies of approximately 60% are achieved, which is remarkable for such small plant sizes. Additionally, heat is also produced to heat the family home when necessary.

© 2013 Elsevier Ltd. All rights reserved.

1. Introduction

High-temperature fuel cells have shown significant potential for use in power production applications. Currently, SOFC systems are the most mature of many options. These systems operate in the range of 750–1000 °C, which makes it possible to combine the SOFC with other conventional thermal cycles to further enhance thermal efficiency. The hybrid SOFC system is considered to be a key technology in achieving future energy goals, based on the many advantages that it offers over other systems.

First, SOFCs do not contain any moving parts. Noise and vibrations during operation are practically non-existent, making it possible to install the system in urban areas. Non-mechanical parts also ensure high reliability and low maintenance costs. Second, SOFCs can operate using a variety of fuels and are more resistant to electrode poisoning. They can tolerate sulfur compounds at concentrations higher than those tolerated by other types of fuel cells. Additionally, unlike in most fuel cells, CO can be used as a fuel in SOFCs. Due to the above-mentioned advantages, SOFCs are considered to be a strong candidate for either hybrid systems or integration into currently deployed technologies.

SOFC-based hybrid power plants have been proposed by researchers, with extensive studies reported on the integration of SOFCs into gas turbine (GT) and steam turbine (ST) power plants, whereas studies on the prospects of combining SOFCs with a Stirling engine are limited. Although SOFC–GT [1–5] and SOFC–ST [6–8] power plants display high efficiencies compared to those of conventional combined cycles, their operations are limited to large-scale plants only. Combined SOFC–Stirling plants offer great potential for small-scale CHP (combined heat and power) generation units, which can be used for houses, small hotels, etc. However, the capital cost of SOFC and Stirling units is currently the primary obstacle in the commercialization of such plants. In this study, the thermodynamic feasibility of a small-scale SOFC–Stirling combined power plant was investigated. Reports of hybrid systems composed of an SOFC system with a Stirling unit can be found in Refs. [9,10]. In Ref. [9], the design possibility of a mobile SOFC–Stirling hybrid plant was studied without presenting the plant design and without analyzing the thermodynamics of the system. A simple study of a 5-kW SOFC/heat engine was presented in Ref. [10], in which SOFC and Stirling engine are connected in parallel for domestic applications.

In this study, a small-scale SOFC–Stirling CHP plant is presented and thermodynamically analyzed. The present study suggests that the SOFC should be placed on top of the Stirling engine to

E-mail address: MR@mek.dtu.dk.

significantly increase the efficiency of the hybrid plant by recovering the off-heat from the SOFC plant, serial connection. Thus, the present study significantly differs from the study presented in Ref. [10], not only in terms of the plant design (parallel versus serial connection) but also based on the thermodynamics of the system. In serial connection, the intention is to increase plant efficiency while this not the case in parallel connection. The details and simulation results concerning plant configurations powered by a variety of fuels are presented. The fuels used in this investigation were natural gas (NG), methanol, ethanol, DME and ammonia. It is shown that the plant configuration with NG is the most complex, requiring the use of a desulfurizer reactor and a pre-reformer reactor. The desulfurizer unit removes the sulfur content of the gas, and the pre-reformer converts the heavier hydrocarbons (hydrocarbons other than methane) to methane, mono-carbon and hydrogen. To feed the plant by methanol, ethanol and DME, it is suggested that a methanator be included to increase the methane content. A methanator is a reactor in which the fuel is reacted with steam and it is suggested that the off-steam available from the SOFC stacks be used. In the majority of previously published studies, the fuel (methanol and DME) was sent directly to the SOFC without the use of a methanator (e.g., [11,12]). Furthermore, it is also demonstrated that the plant design powered by ammonia is the simplest because the fuel can be directly fed to the SOFC cells.

In contrast to most previous studies, which have assumed that natural gas contain only pure methane, in the present study, the gas was a mixture of methane, ethane, propane, carbon dioxide, hydrogen sulfide and *n*-butanes.

The gases after the anode side of the SOFC still possess some fuel, which can be sent to the burner of the Stirling engine. The Stirling engine used in this study is a small-scale (1–5 kW) engine that is currently under development at several companies, such as Stirling Denmark and Stirling Biopower (USA). It is suggested that the engine be cooled by water that is circulated throughout all areas of a house so that the house can be heated via floor heating. The forward water temperature is approximately 40 °C, which is suitable for a family home with floor heating. Alternatively, radiators can be used if the forward temperature of the water can be increased to 45–50 °C. The off-gases produced by the Stirling engine are used to heat the water consumed in the house (water used for showering, washing, etc.).

It should also be noted that the system presented here was studied thermodynamically and that the objective of this study was not to present or discuss the associated costs. The performances of the various plants are compared in terms of efficiency, fuel consumption and other related parameters.

2. Methodology

The results of this study were obtained using the simulation tool dynamic network analysis (DNA), see Ref. [13], which is designed for the analysis of energy systems. This work presents the current results of on-going research in the Department of Mechanical Engineering, Technical University of Denmark, which began as a master's thesis work in [14]. Since then, the DNA program has been developed for general application, it is able to cover unique features, and hence is capable of supplementing other simulation programs.

In DNA, a physical model is formulated by connecting the relevant component models through nodes and by including the operating conditions for the complete system. The physical model is converted to a set of mathematical equations, which are then solved numerically. The mathematical equations include those of mass and energy conservation for all components and nodes as well as relations for the thermodynamic properties of the fluids

involved. At the end of each component, the equations for energy balance and mass balance are included. The total mass balance and energy balance for the entire system are also accounted for. The program is written in FORTRAN.

Throughout the years, a library of different fuels and their respective thermodynamic properties has been compiled, including NG, methanol, ethanol, DME, and ammonia and other fuels.

2.1. Modeling of SOFC stacks

The SOFC model developed in this investigation is based on the planar-type SOFC developed by DTU-Risø and TOPSØE Fuel Cell (TOFC). The model was calibrated against experimental data in the range of 650 °C–800 °C (the operating temperature), as described in Refs. [8,15]. A brief explanation of how the system is implemented in the in-house program is herein provided. The model is assumed to be zero-dimensional, thereby enabling the calculation of complex energy systems. In such modeling tasks, one must distinguish between electrochemical modeling, the calculation of cell irreversibility (cell voltage efficiency) and the species compositions at the outlet. For electrochemical modeling, the operational voltage (E_{cell}) is calculated as follows:

$$E_{\text{cell}} = E_{\text{Nernst}} - \Delta E_{\text{act}} - \Delta E_{\text{ohm}} - \Delta E_{\text{conc}} - \Delta E_{\text{offset}} \quad (1)$$

where E_{Nernst} , ΔE_{act} , ΔE_{ohm} , ΔE_{conc} and ΔE_{offset} are the Nernst ideal reversible voltage, activation polarization, ohmic polarization, concentration polarization and offset polarization, respectively. The contribution of the offset polarization is very small and was therefore neglected in this study. Assuming that only hydrogen is electrochemically converted, the Nernst equation can be written as

$$E_{\text{Nernst}} = \frac{-\Delta g_{\text{f}}^0}{n_e F} + \frac{RT}{n_e F} \ln \left(\frac{p_{\text{H}_2, \text{tot}} \sqrt{p_{\text{O}_2}}}{p_{\text{H}_2\text{O}}} \right), \quad (2)$$

$$p_{\text{H}_2, \text{tot}} = p_{\text{H}_2} + p_{\text{CO}} + 4p_{\text{CH}_4} \quad (3)$$

where Δg_{f}^0 is the Gibbs free energy (for H_2 reaction) at standard pressure, F is the Faraday constant and n_e is the number of charge electrons. R and T are the universal gas constant and the operating temperature, respectively. The water–gas shift reaction is very fast; therefore, the assumption that hydrogen is the only species to be electrochemically converted is justified (see Refs. [16,17]). In the above equations, p_{H_2} and $p_{\text{H}_2\text{O}}$ are the partial pressures for H_2 and H_2O , respectively.

The activation polarization, which is isolated from other polarizations, can be evaluated using the Butler–Volmer equation (see Refs. [18,19]) to determine the charge transfer coefficients and the exchange current density from experimental results by the curve fitting technique. It follows that

$$\Delta E_{\text{act}} = \frac{RT}{(0.001698T - 1.254)F} \sinh^{-1} \left[\frac{i_d}{2(13.087T - 1.096 \times 10^4)} \right], \quad (4)$$

where R , T , F and i_d are the universal gas constant, operating temperature, Faraday constant and current density, respectively.

The ohmic polarization depends on the electrical conductivity of the electrodes as well as the ionic conductivity of the electrolyte, which was also calibrated against experimental data for a cell with anode, electrolyte and cathode thicknesses of 600 μm , 50 μm and 10 μm , respectively. The ohmic polarization can be written as (see, e.g., Refs. [20,21])

$$\Delta E_{\text{ohm}} = \left(\frac{t_{\text{an}}}{\sigma_{\text{an}}} + \frac{t_{\text{el}}}{\sigma_{\text{el}}} + \frac{t_{\text{ca}}}{\sigma_{\text{ca}}} \right) i_{\text{d}}, \quad (5)$$

where $t_{\text{an}} = 600 \mu\text{m}$, $t_{\text{el}} = 50 \mu\text{m}$ and $t_{\text{ca}} = 10 \mu\text{m}$ are the anode, electrolyte and cathode thicknesses, respectively. σ_{an} , σ_{el} and σ_{ca} are the conductivities of the anode, electrolyte and cathode, respectively.

$$\sigma_{\text{an}} = 10^5, \quad \sigma_{\text{ca}} = \frac{5.760 \times 10^7}{T} \exp\left(-\frac{0.117}{8.617 \times 10^{-5} T}\right) \quad (6)$$

$$\sigma_{\text{el}} = 8.588 \times 10^{-8} T^3 - 1.101 \times 10^{-4} T^2 + 0.04679 T - 6.54 \quad (7)$$

The concentration polarization is dominant at high current densities for anode-supported SOFCs, wherein insufficient amounts of reactants are transported to the electrodes and the voltage is thereby reduced significantly. Again, the concentration polarization was calibrated against experimental data by introducing the anode limiting current [22,23], where the anode porosity and tortuosity were also included among the other parameters. The concentration polarization can be modeled as

$$\Delta E_{\text{conc}} = B \left(-\ln\left(1 + \frac{p_{\text{H}_2} i_{\text{d}}}{p_{\text{H}_2\text{O}} i_{\text{as}}}\right) - \ln\left(1 - \frac{i_{\text{d}}}{i_{\text{as}}}\right) \right), \quad (8)$$

where B is the diffusion coefficient, which was calibrated against experimental data and determined to be

$$B = \left(0.008039 X_{\text{H}_2}^{-1} - 0.007272 \right) \frac{T}{T_{\text{ref}}} \quad (9)$$

T_{ref} is the reference temperature (1023 K), and the anode limiting current is defined as

$$i_{\text{as}} = \frac{2 F p_{\text{H}_2} D_{\text{bin}} V_{\text{an}}}{R T_{\text{an}} \tau_{\text{an}}}, \quad (10)$$

where V_{an} and τ_{an} are the porosity and tortuosity of the anode, with values of 30% and 2.5 μm , respectively, in the experimental setup. The binary diffusion coefficient is given by

$$D_{\text{bin}} = \left(-4.107 \times 10^{-5} X_{\text{H}_2} + 8.704 \times 10^{-5} \right) \left(\frac{T}{T_{\text{ref}}} \right)^{1.75} \frac{p_{\text{ref}}}{p}, \quad (11)$$

which was also calibrated against experimental data. p_{ref} is the reference pressure, which was 1.013 bar in this study, and X_{H_2} is the mass reaction rate of H_2 . Finally, the current density i_{d} is directly proportional to the amount of reacting hydrogen according to Faraday's law

$$i_{\text{d}} = \frac{\dot{n}_{\text{H}_2} 2F}{A}, \quad (12)$$

where $\dot{n}_{\text{H}_2}$ is the molar reaction rate of H_2 . The area A is a physical property of the cell and was 144 cm^2 in this study.

The fuel composition at the anode outlet was determined using the Gibbs minimization method, as described in Ref. [24]. Equilibrium at the anode outlet temperature and pressure was assumed for the H_2 , CO , CO_2 , H_2O , CH_4 and N_2 species. Thus, the Gibbs minimization method calculates the compositions of these species at the outlet by minimizing their Gibbs energies. The assumption of equilibrium is justified because the methane content in this study was very low.

To calculate the voltage efficiency of the SOFC cells, the power produced by the SOFC (P_{SOFC}), which depends on the amount of chemical energy fed to the anode, the reversible efficiency (η_{rev}), the voltage efficiency (η_{v}) and the fuel utilization factor (U_{F}), should be taken into account. This power is mathematically defined as follows:

$$P_{\text{SOFC}} = (\text{LHV}_{\text{H}_2} \dot{n}_{\text{H}_2, \text{in}} + \text{LHV}_{\text{CO}} \dot{n}_{\text{CO}, \text{in}} + \text{LHV}_{\text{CH}_4} \dot{n}_{\text{CH}_4, \text{in}}) \eta_{\text{rev}} \eta_{\text{v}} U_{\text{F}}, \quad (13)$$

where U_{F} is a set value and η_{v} is defined as

$$\eta_{\text{v}} = \frac{\Delta E_{\text{cell}}}{E_{\text{Nernst}}} \quad (14)$$

The reversible efficiency is the maximum possible efficiency, defined as the relationship between the maximum electrical energy available (change in Gibbs free energy) and the fuel's LHV (lower heating value), which is detailed as follows (see Ref. [25]):

$$\eta_{\text{rev}} = \frac{(\Delta \bar{g}_{\text{f}})_{\text{fuel}}}{\text{LHV}_{\text{fuel}}} \quad (15)$$

$$\begin{aligned} (\Delta \bar{g}_{\text{f}})_{\text{fuel}} = & \left[(\bar{g}_{\text{f}})_{\text{H}_2\text{O}} - (\bar{g}_{\text{f}})_{\text{H}_2} - \frac{1}{2} (\bar{g}_{\text{f}})_{\text{O}_2} \right] y_{\text{H}_2, \text{in}} \\ & + \left[(\bar{g}_{\text{f}})_{\text{CO}_2} - (\bar{g}_{\text{f}})_{\text{CO}} - \frac{1}{2} (\bar{g}_{\text{f}})_{\text{O}_2} \right] y_{\text{CO}, \text{in}} \\ & + \left[(\bar{g}_{\text{f}})_{\text{CO}_2} + 2 (\bar{g}_{\text{f}})_{\text{H}_2\text{O}} - (\bar{g}_{\text{f}})_{\text{CH}_4} - 2 (\bar{g}_{\text{f}})_{\text{O}_2} \right] y_{\text{CH}_4, \text{in}} \end{aligned} \quad (16)$$

where $\Delta \bar{g}$ is the average Gibbs free energy from the inlet to outlet and y is the molar fraction. The partial pressures were assumed to be the average pressures between the inlet and outlet:

$$\begin{aligned} \bar{p}_j = & \left(\frac{y_{j, \text{out}} + y_{j, \text{in}}}{2} \right) \bar{p} \quad j = \{\text{H}_2, \text{CO}, \text{CH}_4, \text{CO}_2, \text{H}_2\text{O}, \text{N}_2\} \\ \bar{p}_{\text{O}_2} = & \left(\frac{y_{\text{O}_2, \text{out}} + y_{\text{O}_2, \text{in}}}{2} \right) \bar{p}_{\text{c}} \end{aligned} \quad (17)$$

Additionally, the equations for conservation of mass (with molar flows), conservation of energy and conservation of momentum were also incorporated into the model.

The main parameters for the SOFC are presented in Table 1. These parameters should be considered valid unless other values are provided. The number of SOFC stacks was assumed to be 10, and the number of cells per stack was assumed to be 74. The pressure drops in the cathode and anode sides of the SOFC were assumed to be 0.05 and 0.01 bar, respectively. Furthermore, the utilization factor for the SOFC cells and generator (DC/AC) efficiency were assumed to be 0.8 and 0.97, respectively.

Table 1
The main SOFC parameters for the design case.

Fuel utilization factor	0.80
Number of cells in stack	74
Number of stacks	10
Cathode pressure drop ratio (bar)	0.05
Anode pressure drop ratio (bar)	0.01
Cathode inlet temperature ($^{\circ}\text{C}$)	600
Anode inlet temperature ($^{\circ}\text{C}$)	650
Outlet temperatures ($^{\circ}\text{C}$)	780
Generator efficiency	0.97

2.2. Modeling of the Stirling engine

The Stirling engine is noted for its quiet operation and the ease with which it can make use of almost any heat source. Stirling engines are referred to as external combustion heat engines, operated based on a regenerative closed power cycle using helium, nitrogen, air or hydrogen as the working fluid. An ideal regenerative Stirling cycle consists of four processes in one cycle. First, the working fluid absorbs the heat from a high-temperature reservoir and experiences an isothermal expansion (process 1 → 2). Second, the hot working fluid flows through a regenerator, which absorbs heat from the hot working fluid. Thus, the temperature of the working fluid decreases in an isochoric process (2 → 3). Third, the working fluid conducts heat to a low-temperature reservoir and experiences isothermal compression (3 → 4). Finally, the cold working fluid flows back through the regenerator, which conducts heat to the working fluid. The temperature of the working fluid increases in the second isochoric process (4 → 1). In this study, a pseudo Stirling cycle, which more closely approximates actual engine performance, was developed [26]. The main difference between the pseudo Stirling cycle and the ideal Stirling cycle is the assumption of isentropic compression and expansion in the former rather than isothermal compression and expansion in the latter. It is assumed that isentropic compression and expansion provide more realistic cycle performance because, by incorporating these processes, the losses encountered in the Stirling engine are accounted for. In modeling, the engine is divided into three parts: the heater, engine and a cooler. The heater is modeled as a heat exchanger, in which the cold side (the heater wall, T_{hw}) is assumed to have a constant temperature that must be given.

The most important parameters of a Stirling engine are the temperature ratio ζ , the compression ratio ξ , the regenerator effectiveness ε_r and the heater effectiveness ε_h . These are defined as

$$\zeta = \frac{T_1}{T_3}, \quad \xi = \frac{V_4}{V_3}, \quad \varepsilon_r = \frac{T'_2 - T_1}{T_3 - T_1}, \quad \varepsilon_h = \frac{T_{hot,in} - T_{hot,out}}{T_{hot,in} - T_{hw}} \quad (18)$$

where T and V represent temperature and volume, respectively. The subscript 1 indicates the state at the inlet (the working fluid inlet), 2 the state after compression, 3 the state after heating and 4 the state after cooling. State T'_2 corresponds to the state at which the heating process is divided into the isochoric heating of the regenerator and isochoric heating of the heater (see Ref. [26]). The efficiency of such a pseudo Stirling engine can be expressed as

$$\eta_{\text{Stirling}} = \frac{\left[(1 - \xi^{1-\gamma}) - \zeta (\xi^{1-\gamma} - 1) \right]}{\left[(1 - \xi^{1-\gamma}) + (1 - \zeta)(1 - \varepsilon_r) \right]} \quad (19)$$

where γ is the ratio of the specific heats. The work of a Stirling engine is determined from the engine efficiency as

$$W_{\text{Stirling}} = \eta_{\text{Stirling}} (\dot{Q}_{\text{high}} - \dot{Q}_{\text{loss}}) \quad (20)$$

where $\dot{Q}_{\text{high}}$ and $\dot{Q}_{\text{loss}}$ are the heat added by the heater and the heat removed from the engine, respectively. Note that heat is added to and removed from the engine by two different heat exchangers. The losses from the Stirling engine are the result of various loss mechanisms, including mechanical and thermal processes. The heat lost from the engine is estimated by

$$\dot{Q}_{\text{loss}} = (1 - f_{\text{loss}}) \dot{Q}_{\text{high}} \quad (21)$$

where f_{loss} is a “loss factor” that incorporates all mechanisms of heat loss in the engine, including both mechanical and thermal. This loss factor is similar to an efficiency that equals 100% if no losses are present. Such a modeling approach provides reasonable results compared with the experimental data gathered from a Stirling engine. That is, such modeling is simple but still yields reasonable results if f_{loss} is calibrated using performance data.

In Eq. (18), T_1 is the lowest temperature (the cooler wall temperature, T_{cw}), and T_3 is the highest temperature (working fluid temperature inside the engine, T_{hg}). These temperatures are calculated by

$$T_{hg} = T_{hw} - \Delta T_{\text{heater}}, \quad T_{cw} = T_{cw,in} + \frac{2}{3} \Delta T_{\text{cooler}} \quad (22)$$

Thus, the highest temperature of the working fluid (helium) is lower than the heater wall temperature, and the lowest temperature of the working fluid is a weighted temperature and is the average between the inlet and outlet temperatures. ΔT_{heater} and ΔT_{cooler} refer to the temperature difference over the heater and cooler, respectively. The main parameters for the Stirling engine are presented in Table 2.

2.3. Modeling of the pre-reformer and methanator

To model the pre-reformer and methanator, a simple Gibbs reactor is implemented [8,24], meaning that the total Gibbs free energy attains a minimum when chemical equilibrium is achieved. Such a characteristic can be used to calculate the outlet gas composition at a specified temperature and pressure without considering the reaction pathways. However, the species at the outlet must be specified, which can vary from the methanator to the pre-reformer. The species at the outlet of the pre-reformer are defined as H_2 , CO , CO_2 , steam, CH_4 , N_2 , NO , H_2S , SO_2 , NO_2 , NH_3 and Ar. For the methanator, the species at the outlet are defined as H_2 , CO , CO_2 , steam, CH_4 , N_2 , NO , H_2S , SO_2 , NO_2 , HCN (hydrogen cyanide), COS (carbonyl sulfide), N_2O , NO_3 , SO_3 and Ar.

2.4. Modeling the other components

The power consumption of the pumps was calculated as

$$W_{\text{pump}} = \left[\frac{\dot{m} v_{\text{in}} (p_{\text{out}} - p_{\text{in}})}{\eta} \right]_{\text{pump}} \quad (23)$$

where $\dot{m}$, p , v and η are the mass flow, pressure, specific volume (m^3/kg) and efficiency of the pump, respectively. The pump efficiency and outlet pressure were defined as shown below. The power consumption of the compressors was modeled based on the definitions of isentropic and mechanical efficiencies (given values) as follows:

Table 2

The main Stirling engine parameters for this design case.

Heater and cooler Δp (bar)	0.01
Heater wall temperature, ($^{\circ}C$)	600
Heater ΔT ($^{\circ}C$)	125
Heater effectiveness	0.95
Cooler ΔT ($^{\circ}C$)	60
Compressor ratio (—)	1.44
Regenerator effectiveness	0.98
Loss factor	0.8

Table 3

The main parameters for the accessory components.

Compressor intake temperature (°C)	25
Compressor isentropic efficiency	0.7
Compressor mechanical efficiency	0.95
Fuel side heat exchangers Δp (bar)	0.01
Air/Gas side heat exchangers Δp (bar)	0.05
Water side heat exchangers Δp (bar)	0.3
Desulfurizer temperature (°C)	200
Depleted air temperature (°C)	40
Depleted exhaust gas temperature (°C)	95
Fuel inlet temperature (°C)	25

$$\eta_{is} = \left[\frac{h_{out, \text{Sin}} - h_{in}}{h_{out} - h_{in}} \right]_{\text{compressor}} \quad (24)$$

$$\eta_m = \left[\frac{\dot{m}(h_{out} - h_{in})}{W} \right]_{\text{compressor}} \quad (25)$$

where h is the enthalpy, h_s is the enthalpy when the entropy is constant. The subscripts in and out refer to the inlet and outlet of the component.

In modeling the heat exchanger, it was assumed that all energy from one side is transferred to the other side by neglecting the heat losses. Depending on the type of heat exchanger used, both the LMTD (logarithmic mean temperature difference) and ϵ -NTU (effectiveness-number of transferred unit) methods were used (see Ref. [27]).

The desulfurizer unit is a simple model in which the sulfur content is removed. The compositions are re-calculated after sulfur removal. The main parameters for the accessory components are presented in Table 3. The pressure drops for all heat exchangers are assumed to be 0.01 bar at the fuel side, 0.05 bar at the air side and 0.3 bar at the water side. Because the system is designed for low-scale power, the fuel and air mass flows tend to be small, resulting in lower efficiencies of the turbomachines. Therefore, the compressor isentropic efficiency and mechanical efficiency are assumed to be 0.7 and 0.95, respectively. The depleted air temperature is set to 40 °C, whereas the depleted exhaust gas temperature is assumed to be 95 °C.

3. Plant configurations

The proposed combined power system is composed of SOFC and Stirling systems connected in series. The fuel is fed to the topping SOFC cycle only, where the fuel reforming and electrochemical oxidation processes occur. The SOFC stack produces electrical power in addition to an exhaust stream that contains unused CO and H₂. The unburned fuels from the anode side combined with the exhaust air from the cathode side are sent immediately to a catalytic burner for further combustion.

The plant simulations were conducted using different fuels, including natural gas (NG), DME, ethanol, methanol and ammonia. The configuration of the SOFC plant varies depending on the nature of the fuel. In contrast, the Stirling plant retains the same configuration because it is operated by only the heat generated from the exhaust gases in the burner. Three configurations are suggested for the combined SOFC–Stirling system: the first configuration uses natural gas as the fuel, as shown in Fig. 1. The fuel is preheated to 200 °C in a heat exchanger before it is sent to a desulfurization unit to remove the sulfur content from the NG. In most previous studies, NG has been assumed to contain only CH₄ (see Refs. [28–30]), while in the present study, the NG was a mixture of lighter and heavier hydrocarbons, including hydrogen sulfide (see Table 4).

As is known, SOFCs are capable of the direct internal reformation of light hydrocarbons but not heavier hydrocarbons (such as methane) at the anode. Using high operating temperatures, the methane can be directly reformed to H₂ and CO. Due to the small amount of heavier hydrocarbons present in NG, an external reforming system is employed to avoid the thermal shock caused by the endothermic reforming reaction.

The heavier hydrocarbons in the NG are reformed in a CPO (catalytic partial oxidation)-type pre-reformer reactor, which was detailed in the preceding section. Another heat exchanger heats the fuel to the designated temperature of the pre-reformer. Due to the exothermic nature of the CPO reactions, the outlet temperature is high enough to support direct feeding into the SOFC. However, for the sake of security, a heat exchanger is included after the pre-reformer to ensure that the pre-reformed fuel reaches a temperature of at least 650 °C before entering the anode side of the SOFC. Including the an anode preheater (AP) would be beneficial during

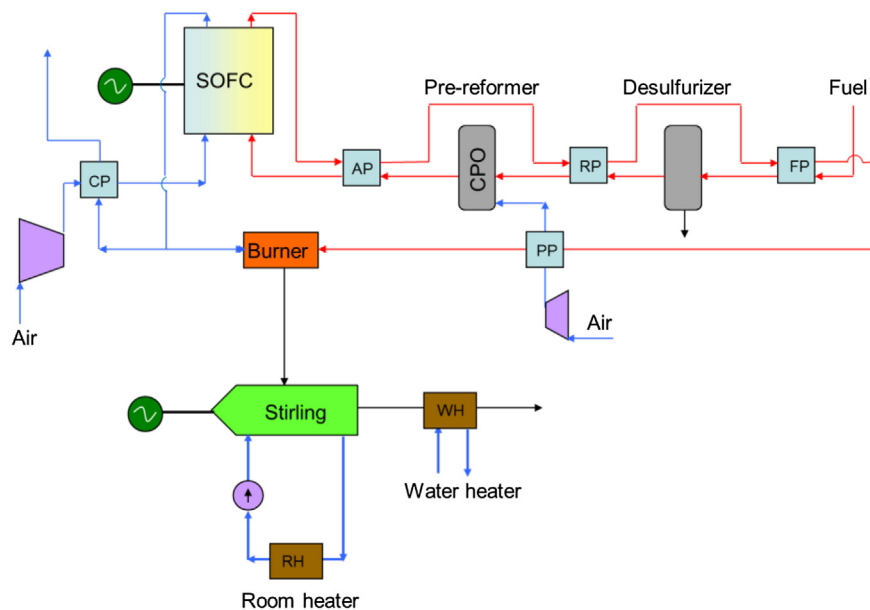
**Fig. 1.** A schematic of the combined SOFC–Stirling system fueled by natural gas.

Table 4

The molar fraction of the natural gas used in the present study.

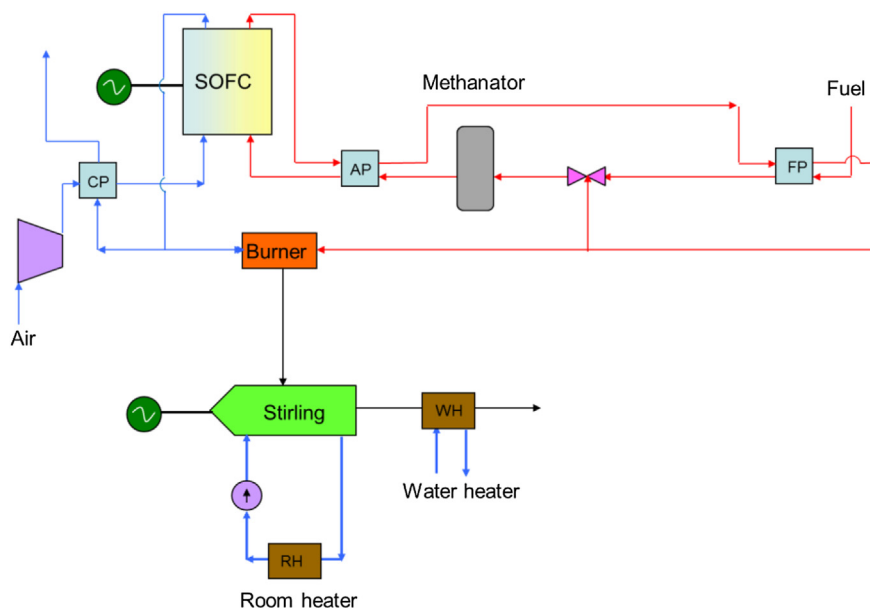
Molecular formula	Chemical name	Molar fraction %
CH ₄	Methane	87
C ₂ H ₆	Ethane	8.1
C ₃ H ₈	Propane	1
CO ₂	Carbon dioxide	2.925
H ₂ S	Hydrogen sulfide	0.375
C ₄ H _{10–n}	<i>n</i> -Butane	0.6

the start-up sequence. The operating temperature of the SOFC stack is assumed to be 780 °C, which is also assumed to be the outlet temperature of the SOFC cells, for both the anode and cathode sides. The thermal energy of the exhaust fuel is recuperated for to heat the incoming fuel at all of the stages mentioned above.

On the cathode side of the SOFC, air is compressed in a compressor and then preheated in a heat exchanger (cathode preheater, CP) to approximately 650 °C. The electrochemical reactions at the anode and cathode produce electrical power and heat at the expense of the fuel. Part of the released heat is utilized in the endothermic reforming process. The off-air produced beyond the cathode side is divided into two streams, one for the cathode preheater and the other for the burner. This design ensures a sufficiently high temperature for the burner, which provides the necessary heat for the bottoming cycle in the Stirling engine. Thus, the current design differs slightly from that proposed in Ref. [7].

The depleted fuel still contains some unutilized combustible fuel, such as hydrogen and carbon monoxide, which is sent to a burner to generate heat for the bottoming Stirling cycle. To remove heat from the Stirling engine and cool it down, a heat exchanger is used. The other side of which heats all spaces in the house. This heat exchanger is designated as RH (the room heater) in the figures. Thus, the burner acts as a heat source for the Stirling engine, whereas the RH acts as heat sink for the Stirling engine. The exhaust gas following the Stirling engine still possesses valuable heat, which can be used to warm the water consumed in the house, such as that used for showering and washing. This heat exchanger is designated WH (water heater) in the figures.

A schematic of the second configuration is shown in Fig. 2. DME, ethanol and methanol are used as the fuels in this configuration.

**Fig. 2.** A schematic of the combined SOFC–Stirling system with methanol, DME and ethanol as fuels.

Unlike natural gas, these fuels do not require any desulfurization unit. Additionally, a methanator is used to crack the fuel, increasing the amount of methane.

Due to the endothermic nature of methane reforming inside the SOFC stacks, the need for excess air to cool the SOFC stacks is reduced. In turn, the work performed by the air compressor is reduced, thereby increasing the plant efficiency. The fuel is preheated before reaching the methanator, where it is reformed to CH₄, H₂ and CO. A heat exchanger heats the fuel to the desired temperature for the SOFC inlet. A portion of the exhaust fuel from the SOFC is recycled and mixed with the fuel at the inlet of the methanator to provide the needed steam for reactions in the methanator. For the high-temperature recycle pump, long-term durability is a problem. However, for the low-temperature pump, the off-fuel stream must be cooled to a low temperature, which results the waste of substantial thermal energy. Thus, in this study, an ejector was incorporated into the simulations. The rest of the plant configuration remained the same as that in the NG case.

A schematic of the third configuration is shown in Fig. 3, in which ammonia is used as the fuel. Because ammonia can be directly fed to the SOFC cells, unlike in the NG case, neither a desulfurization reactor nor a pre-reformer reactor is needed. It is sufficient to preheat the ammonia to the desired inlet temperature of the SOFC anode (i.e., 650 °C). The rest of the system remains the same as that discussed in the previous cases.

3.1. Fuel reforming

Although it has been shown that the direct cracking of hydrocarbons in SOFCs is possible with internal reforming, the effect of thermal shock, which occurs as a result of endothermic reactions, on SOFC materials becomes a limiting factor. For small-scale SOFC applications, fuel reforming and adequate preheating of the inlet flows must be performed efficiently. In the current investigation, a CPO (catalytic partial oxidation) reactor was selected because of its rapid kinetics and the exothermic nature of its operation. When compared to an adiabatic steam reformer (ASR), the CPO provides better start-up and transient responses and reduces the heat input to the fuel stream, both of which are of significant importance in small-scale power plants. For the pre-reformer, the fuel inlet

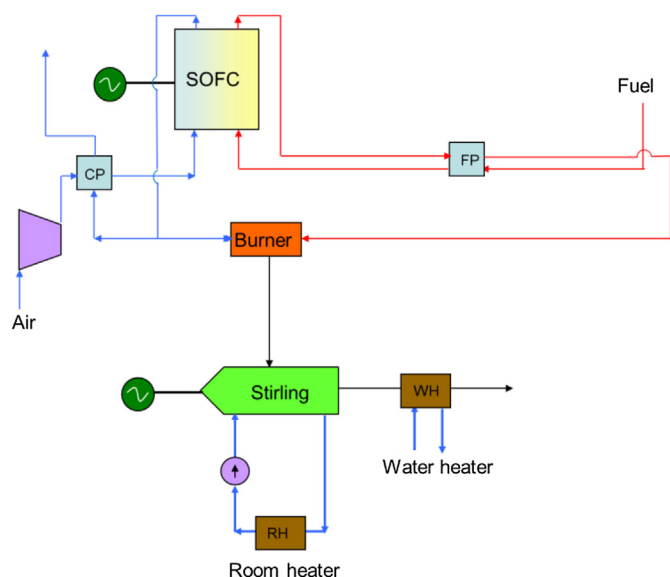


Fig. 3. A schematic of the combined SOFC–Stirling system fueled by ammonia.

temperature must be lower than 400 °C and the outlet temperature should be over 500 °C to prevent carbon coking and to achieve a certain level of hydrocarbon pre-reforming. In the above-mentioned configuration for NG, waste heat from the anode exhaust provides preheated fuel temperatures of 300 °C or above for the CPO reactor.

The outlet temperature of the stream is restricted to 550 °C to maintain a specific O/C ratio. Therefore, an additional heat exchanger for fuel preheating is introduced before the SOFC. Conversely, systems with CPO pre-reformers do not require additional fuel preheating if the O/C ratio is maintained above 1. For DME, ethanol and methanol, a methanator is used, which inadvertently operates like an ASR. The off-fuel is partially recycled and then mixed with the incoming fuel. The inlet fuel temperatures and the off-fuel recirculation depend on the O/C ratio in the reformer. Unlike in an ASR, the reactions for DME and ethanol in the methanator are exothermic. To prevent carbon coking, a specific O/C ratio value should be selected for each fuel type, e.g., 1.7 for NG and 1.5 for DME, ethanol and methanol. Alternatively, the outlet temperature of the reactor must also be specified. Because the off-fuel includes CO₂, H₂ and CO in addition to H₂O, the ratio (H + O)/C should be considered instead of S/C [31], which was considered in this study.

4. Results and discussion

It is important to mention that the net power output of the plant was fixed to 10 kW for all fuels. All power inputs and outputs, species flows, heat losses and heat sources were balanced in the simulations, making the modeled plants thermally self-sustainable, i.e., no extra heat or fuel was needed. Fuel was fed only to the SOFC cycle. Because the burner in the Stirling engine extracts fuel from the SOFC exhaust stream, the fuel feed remains the same for both the plants. The most important results are presented and discussed below. The main calculated parameters are shown in Table 5. The plant efficiencies shown in the table are based on LHV only.

One of the notable differences between the configurations is the air consumption: methanol uses more air to produce 10 kW of electricity than other fuels, although NG has the highest air-fuel ratio. Higher air mass flow leads to increased power consumption. However, the least amount of air is compressed for the SOFC

Table 5

The calculated parameters for the hybrid plant.

Parameter	NG	DME	Ethanol	Methanol	Ammonia
Fuel mass flow (kg/h)	1.33	2.13	2.15	3.06	3.34
Air compressor mass flow (kg/h)	58.44	60.82	52.09	71.00	58.71
Air-fuel ratio (–)	43.9	28.6	24.2	23.2	17.6
CPO or methanator inlet T (°C)	525	300	300	300	–
CPO or methanator outlet T (°C)	650	595	528	566	–
SOFC off-fuel (kg/h)	6.44	5.87	5.94	6.98	7.02
Fuel recirculated to methanator (%)	–	2.8	3.2	3.3	–
Burner temperature (°C)	1256.0	1304.8	1346.9	1200.1	1152.2
Stirling outlet temperature (°C)	632.8	635.2	637.4	630.0	627.6
Stirling power (kW)	1.105	1.177	1.121	1.211	1.062
Power consumption (kW)	0.131	0.135	0.116	0.158	0.135
Thermal efficiency (LHV) (%)	59.03	58.58	62.62	59.04	57.89

when ethanol is fed to the system. Only a small amount of air is compressed is because reformed ethanol contains high levels of methane (CH₄), which leads to more internal reforming (endothermic reactions) in SOFCs in comparison to other fuels, and consequently, less air is needed to cool the stack. Moreover, in contrast to NG, higher percentages of water in the reformed ethanol promote endothermic reforming in the fuel cell.

It also can be observed that the methanator temperature (at the outlet) is the lowest for the case in which ethanol is the fuel. Thus, the least amount of CO is produced in the methanator. It should be noted that in the NG system, a CPO reformer is used and not a methanator. Therefore, the conditions for the reformer will also be different.

The burner temperature for the ethanol case is the highest, whereas the power produced from the Stirling engine is the highest for the case with methanol. The reason for this result can be explained by the fuel mass flow through the burner, which in turn depends on how much fuel has been re-circulated (if any) for the corresponding fuel type; see Fig. 2. However, the power from the Stirling engine is approximately similar for all fuel types considered and changes within 10% from the highest to the lowest case.

The power consumed depends mainly on the air compressor, which in turn depends on the air mass flow through the component. As the mass flow increases, more power is consumed. Table 5 shows that for the methanol case, the power consumption is the highest because more air is compressed to maintain the SOFC temperature at the desired level.

The plant efficiency for the ethanol case is the highest, whereas the ammonia plant has the lowest efficiency.

4.1. Effect of utilization factor

The results presented in Table 5 are for systems with a fuel cell utilization factor of 0.8. Running the SOFCs with different fuel utilization factors affects the plant efficiency, as shown in Fig. 4. It should be noted that the fuel mass flow remains unchanged when the fuel utilization factor is varied. The net output power changes insignificantly because the power produced from the topping cycle decreases slightly, while the power produced from the bottoming cycle is simultaneously increased (as discussed in Refs. [32,33]). This behavior occurs because decreasing the fuel utilization factor results in more fuel fed to the burner; therefore, more energy is produced in the bottoming cycle.

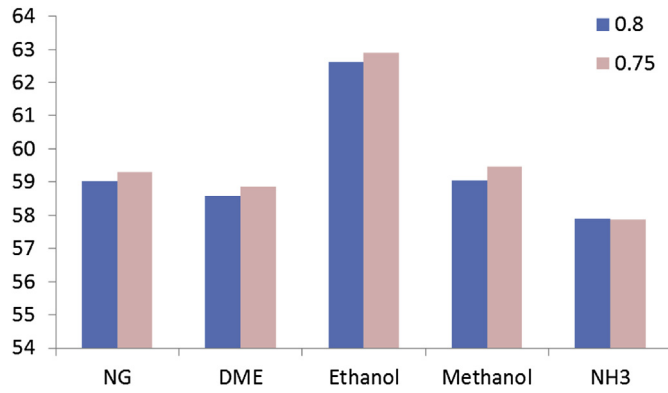


Fig. 4. The effect of the SOFC fuel utilization factor on plant efficiency.

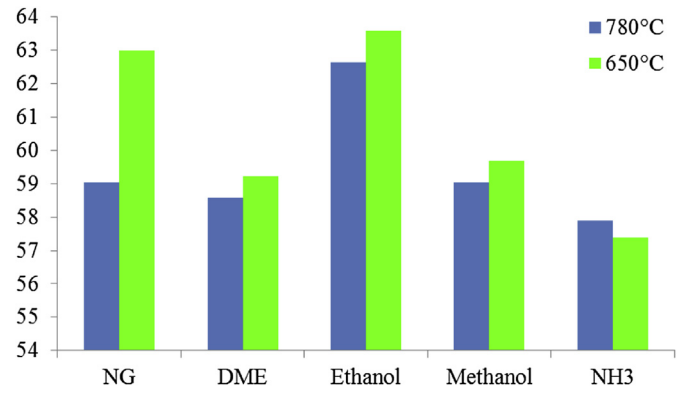


Fig. 5. The effect of SOFC operating temperature on plant efficiency.

Thus, for these hybrid systems lowering the utilization factor results in increased plant efficiency. However, the fuel cells cannot be run with utilization factors that are too low because the investment cost of the SOFC stacks is relatively high, and it would not be economical to run the stacks with utilization factors less than approximately 0.75. Thus, instead of lowering the utilization factor, one can decrease the number of cells in the stack; in contrast, the fuel cells can be run with higher utilization factors to reach the same net output power, which is more economical. As shown, the effect is least pronounced in the case for ammonia fuel.

4.2. Effect of the SOFC operating temperature

Some SOFC manufacturers, such as the manufacturer of the TOPSØE Fuel Cell, are trying to decrease the operating temperature of the cells, and a target value of 650 °C has been indicated for future generations of the technology. Therefore, it is of interest to study the plant behavior at reduced operating temperatures. Decreasing the SOFC operating temperature allows for the inlet temperatures of the anode and cathode to also be decreased. Thus, in the simulation, these temperatures were decreased by 130 °C, accordingly. This assumption is reasonable because no information is available concerning the inlet temperatures for future SOFC generation. Lowering the anode inlet temperature by such an extent allows for the removal of the anode preheater from the system when the plant is fed by methanol, ethanol or DME (AP in Fig. 2). Thus, all plants would then be able to run without any anode preheater, and the fuel after the reformer/methanator would be fed directly to the anode of the SOFCs.

Other factors worth discussing are the operating temperatures of the NG pre-reformer and methanator. As mentioned previously, the outlet temperature of the pre-reformer running on NG is sufficiently high to feed the reformed gas directly to the anode side of the fuel cell, without incorporating any anode preheater. Decreasing the anode temperature allows the inlet temperature of the pre-reformer to be decreased as well. Thus, this temperature is lowered by 130 °C–495 °C, which is still sufficiently high to carry out a complete pre-reforming process. The inlet temperature of the methanator would still be the same as that in the previous case, i.e., 300 °C, at which the methanation reaction is complete (see Ref. [34]).

Considering the temperature levels mentioned above, the simulation results for the SOFC operating temperature of 650 °C are presented in Fig. 5. As shown, the plant efficiency increases for all fuels except ammonia. It should be noted that the fuel mass flow was unchanged and that the polarization losses were assumed to be the same as those in the previous cases because no information is available for the new generation of SOFCs.

For methanol, ethanol and DME, the increase in plant efficiency can be explained by the fact that the anode preheater has been removed; hence, no energy is used to preheat the reformed fuel prior to the anode inlet of the SOFC stacks. However, less air mass flow is required to cool the SOFC stacks, which are now lowered. As discussed in Refs. [1,8], the reason that the air mass flow is significantly higher than the stoichiometric value is to cool the SOFC stacks, maintaining their operating temperature at the desired value. The required excess air depends entirely on the nature of the fuel entering the stacks. Thus, the power consumed by the compressor is significantly decreased. Consequently, the burner temperature is increased and more energy is available for the bottoming cycle in the Stirling engine. The power produced by the SOFC stacks is decreased, but the power produced by the Stirling engine is increased.

For the case with NG, the pre-reformer inlet temperature is decreased, which means that less energy is used to preheat the fuel. Furthermore, less air mass flow is required to cool the SOFC stacks, which decreases the power consumption of the compressor (22% less). Additionally, the power consumed by the reformer compressor is also decreased (by approximately 66%). Consequently, the plant efficiency is increased to a greater degree than in the other cases.

The plant using ammonia does not possess a reformer, which means that decreasing the operating temperature of the SOFC directly decreases the burner temperature. Thus, less energy is available for the bottoming cycle, and consequently, the plant efficiency decreases. However, the decrease is small, i.e., from 57.9% to 57.4%.

4.3. Additional considerations

It has been shown [7,8] that when two plants are serially integrated, the hybrid plant efficiency can be calculated by

$$\eta_{\text{plant}} = \eta_{\text{top cycle}} + \eta_{\text{bottom cycle}} (1 - \eta_{\text{top cycle}}) \varepsilon \quad (26)$$

where ε is the effectiveness of the heat exchanger that transfers heat from the topping cycle to the bottoming cycle. In the current study, the topping cycle was the SOFC plant, and the bottoming cycle was the Stirling engine. Thus, Eq. (26) can be re-written as

$$\eta_{\text{plant}} = \eta_{\text{SOFC}} + \eta_{\text{Stirling}} (1 - \eta_{\text{SOFC}}) \varepsilon_{\text{SOFC to Stirling}} \quad (27)$$

It can be shown that ε can be related to the temperatures when a heat recovery steam generator (HRSG) is used to transfer heat from the topping cycle to the bottoming cycle (see Refs. [7,8,32]). However, in the current design, no HRSG is used;

Table 6

The accuracy of Eqs. (27) and (28).

Efficiency %	NG	DME	Ethanol	Methanol	Ammonia
SOFC plant	52.5	51.7	55.6	51.9	51.8
Stirling engine plant	26.1	26.1	26.1	26.1	26.1
Hybrid plant	59.0	58.6	62.6	59.0	57.9
Efficiency by Eq. (27)	58.9	58.4	61.9	58.1	57.7
Effectiveness by Eq. (28)	51.2	52.9	54.3	49.1	47.2

therefore, ε must be defined differently. It has been proposed that ε be defined as

$$\varepsilon_{\text{SOFC to Stirling}} = \frac{T_{\text{Stirling, in}} - T_{\text{Stirling, out}}}{T_{\text{Stirling, in}} - T_{\text{water, out}}} \quad (28)$$

which can be explained as the heat from the burner entering the Stirling engine at the temperature of $T_{\text{Stirling, in}}$ and leaving the engine at $T_{\text{Stirling, out}}$. That is, not all of the heat is used in the engine. The lowest temperature at which the heat from the burner can be used in the engine is the same as that of the cooling water leaving the engine, not the ambient temperature. Therefore, $T_{\text{water, out}}$ is used as the lowest temperature in Eq. (28). This equation facilitates the characterization of hybrid plants with high accuracy when the efficiency of each plant is known. Table 6 demonstrates the accuracy of Eqs. (27) and (28). The power consumption of the water pump was considered for the Stirling engine plant when calculating its efficiency.

As shown in Table 6, Eqs. (27) and (28) can predict the hybrid plant efficiency with very high accuracy if the efficiency of each cycle (i.e., the SOFC plant and the Stirling engine) combined with ε (the temperatures around the Stirling engine, Eq. (28)) is known. The differences between the simulations and Eq. (27) are less than 2% for all cases considered. For example, the simulated hybrid plant efficiency fed by NG is 59%, whereas Eq. (27) gives a value of 58.9%. The table also shows that the Stirling plant had an efficiency of approximately 26% regardless of the fuel type. However, the efficiency of the SOFC plant fed by ethanol was the highest; therefore, the hybrid plant efficiency for this system was also the highest. In contrast, the effectiveness (Eq. (28)) was the lowest when the plant was fed by ammonia, which also explains why the hybrid plant fed by ammonia showed the lowest value.

Due to the lack of experimental data, the results obtained were further compared with those reported in other studies in the literature. For NG, the results obtained can be compared with those of the study reported in Ref. [11], in which the Aspen Plus software program was used for the simulations; see Table 7. Satin et al. [11] obtained 51.3% efficiency for the SOFC plant alone, which is in good agreement with the result obtained in this study, i.e., 52.5%. Using methanol as the fuel, the results obtained in this study can be compared with those of the studies reported in Refs. [35,36]. In [35], an efficiency of approximately 50% was simulated for the SOFC plant alone, whereas in Ref. [36], the value was 53.7%. For the

Table 7

A comparison of the SOFC plant efficiency simulated in this study with the results presented in the literature.

	NG	DME	Ethanol	Methanol	Ammonia
Current study	52.5	51.7	55.6	51.9	51.8
Literature [11]	51.3				
Literature [12]		~50.1			
Literature [35]				~50	
Literature [36]			55.6	53.7	
Literature [37]			~55		
Literature [38]					~51

ethanol case, the results of Ref. [36] are exactly the same as those of the current study, which also yielded 55.6%. Ammonia was fed to an SOFC in the study presented in Ref. [38] with a different current density. Interpolating the results to 300 mA (close to the highest current used in the current study), a plant efficiency of approximately 51% is obtained, which is also close to the result obtained here, i.e., 51.8%.

For the Stirling engine case, the results obtained in this study agree well with those for the Stirling engine reviewed in Ref. [39], in which two commercially available engine are mentioned, one with an efficiency of 18% (Stirling Denmark, SD-1) and the other one with an efficiency of 30% (Stirling Biopower). Here, an efficiency of approximately 26% was calculated.

5. Cost estimation

A detailed economic analysis is not within the scope of this study, but to estimate the plant cost, one can assume the cost of the fuel cell stacks (C_{SOFC}) based on Ref. [40] to be

$$C_{\text{SOFC}} = (N_{\text{cell}} A_{\text{cell}})(2.96T_{\text{cell}} - 1907) \quad (29)$$

where N_{cell} , A_{cell} and T_{cell} are the number of cells, the cell area and the cell temperature, respectively. It should be noted that this equation is valid for the mass production of SOFC stacks. Using the data presented in Table 1, the cost of an SOFC stack can be estimated to be approximately \$4280, which is valid for approximately 8.8 kW of electricity. It was discussed in Ref. [41] that the cost of the SOFC stacks is approximately one third of the SOFC plant cost. Thus, the cost of the SOFC plant alone, with all related instrumentation and control systems, can be estimated to be \$12,840, which is also valid for 8.8 kW of electricity. Including the installation cost (an additional 35% [40]), the total cost of the SOFC plant is estimated to be \$17,330.

The cost of a Stirling engine with a 1.2-kW power output can be estimated to be \$3300 (see Ref. [39]). Thus, the total cost of the hybrid plant presented here can be roughly approximated to be \$20,630 (approximately 15,900 €), which is valid for a 10-kW plant. As mentioned above, this cost is a rough approximation; to calculate the cost in detail, one must study the costs of each component in addition to the installation cost and the investment cost and payback time. Then, from the payback time, one can select the size of plant desired for purchase, ranging from 1 kW to 10 kW. Such a study is very comprehensive and beyond the scope of this paper.

6. Conclusion

A novel hybrid SOFC–Stirling plant with a power capacity of up to 10 kW was designed and presented with the relevant thermodynamic calculations.

- A plant efficiency of approximately 60% was achieved, which is remarkable for such a small plant.
- With slight decrease in fuel utilization factor of the SOFC, fuel supply to the burner increases, and the plant efficiency enhances regardless of fuel used since power produced from the bottom cycle of Stirling engine was found increasing.
- By lowering SOFC operating temperature, the plant efficiency will be decreased for all fuels except with ammonia. The increase in plant efficiency fueled by natural gas was more pronounced than other fuels; methanol, ethanol and DME.
- The SOFC plant efficiencies calculated from mathematical model and compared with earlier studies are found in good agreement.

- v) The efficiencies are also found to be close to the engine of the manufacturers presented here which validates the authenticity of the mathematical model.
- vi) The estimated cost of such hybrid plant was also found approximately to 2060 \$/kW (approx. 1580 €/kW) on the basis of simple cost analysis.

References

- [1] Rokni M. Introduction of a fuel cell into combined cycle: a competitive choice for future cogeneration. In ASME Cogen – Turbo IGTI 1993; vol. 8. p. 255–61.
- [2] Calise F, Dentice d'Accadia M, Palombo A, Vanoli L. Simulation and exergy analysis of a hybrid solid oxide fuel cell (SOFC)–gas turbine system. Energy 2006;31:3278–99.
- [3] Komatsu Y, Kimijima S, Szmyd JS. Performance analysis for the part-load operation of a solid oxide fuel cell–micro gas turbine hybrid system. Energy 2010;35:982–8.
- [4] Roberts RA, Brouwer J. Dynamic simulation of a pressurized 220 kW solid oxide fuel cell–gas turbine hybrid system: modeled performance compared to measured results. Fuel Cell Science and Technology 2006;3(1):18–25.
- [5] Bang-Møller C, Rokni M, Elmegaard B. Exergy analysis of biomass gasification, solid oxide fuel cell and micro gas turbine hybrid systems. Energy 2011;36:4740–52.
- [6] Dunbar WR, Lior N, Gaggioli RA. Combining fuel cells with fuel fired power plants for improved exergy efficiency. Energy 1991;16(10):1259–74.
- [7] Rokni M. Thermodynamic analysis of an integrated solid oxide fuel cell cycle with a Rankine cycle. Energy Conversion and Management 2010;51(12):2724–32.
- [8] Rokni M. Plant characteristics of an integrated solid oxide fuel cell and a steam cycle. Energy 2010;35:4691–9.
- [9] Winkler W, Lorenz H. Design studies of mobile applications with SOFC-heat engine models. Journal of Power Sources 2002;106:338–43.
- [10] Foley AC. A unique sub 5 kW solid oxide fuel cell/heat engine hybrid generator. Papers in American Chemical Society 2004;49(2):789–91.
- [11] Santin M, Traverso A, Magistri L, Massardo A. Thermoeconomic analysis of SOFC-GT hybrid systems fed by liquid fuels. Energy 2010;35:1077–83.
- [12] Cocco D, Tola V. Externally reformed solid oxide fuel cell–micro-gas turbine (SOFC–MGT) hybrid systems fueled by methanol and di-methyl-ether (DME). Energy 2009;34:2124–30.
- [13] Elmegaard B, Houbak N. DNA – a general energy system simulation tool. In: Proceeding of SIMS, Trondheim, Norway 2005.
- [14] Perstrup C. Analysis of power plant installation based on network theory. M.Sc. thesis. Laboratory of Energetics, Denmark: Technical University of Denmark; 1989 (in Danish).
- [15] Petersen TF, Houbak N, Elmegaard B. A zero-dimensional model of a 2nd generation planar SOFC with calibrated parameters. International Journal of Thermodynamics 2006;9(4):161–8.
- [16] Holtappels P, DeHaart LGJ, Stimming U, Vinke IC, Mogensen M. Reaction of CO/CO₂ gas mixtures on Ni–YSZ cermet electrode. Applied Electrochemistry 1999;29:561–8.
- [17] Matsuzaki Y, Yasuda I. Electrochemical oxidation of H₂ and CO in a H₂–H₂O–CO–CO₂ system at the interface of a Ni–YSZ cermet electrode and YSZ electrolyte. Journal of Electrochemical Society 2000;147(5):1630–5.
- [18] Keegan KM, Khaleel M, Chick LA, Recknagle K, Simner SP, Diebler J. Analysis of a planar solid oxide fuel cell based automotive auxiliary power unit. SAE technical paper series No. 2002-01-0413; 2002.
- [19] Prentice G. Electrochemical engineering principles. Houston, USA: Prentice Hall International; 1991.
- [20] Achenbach E. Three-dimensional and time-dependent simulation of a planar solid oxide fuel cell stack. Journal of Power Sources 1994;49(1–3):333–48.
- [21] Zhu H, Kee RJ. A general mathematical model for analyzing the performance of fuel-cell membrane-electrode assemblies. Journal of Power Sources 2003;117:61–74.
- [22] Costamagna P, Selimovic A, Del Borghi M, Agnew G. Electrochemical model of the integrated planar solid oxide fuel cell (IP-SOFC). Chemical Engineering 2004;102(1):61–9.
- [23] Kim JW, Virkar AV. The effect of anode thickness on the performance of anode-supported solid oxide fuel cell. In: Proceedings of the sixth international symposium on SOFCs, (SOFC-VI), PV99-19. The Electrochemical Society; 1999. p. 830–9.
- [24] Smith JM, Van Ness HC, Abbott MM. Introduction to chemical engineering thermodynamics. 7th ed. Boston: McGraw-Hill; 2005.
- [25] Winnick J. Chemical engineering thermodynamics. New York: John Wiley & Sons; 1997.
- [26] Reader GT. The Pseudo Stirling Cycle – a suitable performance criterion. In Proceeding of the 13th intersociety energy conversion engineering conference August 20–25, 1979; vol. 3. p. 1763–70. San Diego, California.
- [27] Incropera FP, DeWitt DP, Bergman TL, Lavine AS. Introduction to heat transfer. 5th ed. Wiley, ISBN 978-0471457275; 2006.
- [28] Kandepu R, Imsland L, Foss BA, Stiller C, Thorud B, Bolland O. Modeling and control of a SOFC-GT-based autonomous power system. Energy 2007;32:406–17.
- [29] Jia J, Li Q, Luo M, Wei L, Abdudula A. Effect of gas recycle on performance of solid oxide fuel cell power system. Energy 2011;36:1068–75.
- [30] Al-Sulaiman FA, Hamdullahpur F, Dincer I. Performance of three trigeneration system using organic Rankine cycles. Energy 2011;36:5741–54.
- [31] Sasaki K, Batteries Teraoka Y. Fuel cells, and energy conversion – equilibria in fuel cell gases – II. The C–H–O ternary diagram. Electrochemical Society 2003;150(7):A885–8.
- [32] Kehlhofer RH, Warner J, Nielsen H, Bachmann R. Combined – cycle gas steam turbine power plants. 2nd ed. Oklahoma: PennWell, ISBN 0-87814-736-5; 1999.
- [33] Rokni M. Thermodynamic investigation of an integrated gasification plant with solid oxide fuel cell and steam cycles. Green 2012;2:71–86.
- [34] Liu H, Zhang J. Electrocatalysts of direct Methanol fuel cell: from fundamentals to applications. Wiley-VCH, ISBN 978-3-527-32377-7; 2009.
- [35] Doherty W, Reynolds A, Kennedy D. Computer simulation of a biomass gasification-solid oxide fuel cell power system using Aspen Plus. Energy 2010;35:4545–55.
- [36] Leone P, Lanzini A, Ortigoza-Villalba GA, Borchellini R. Operation of a solid oxide fuel cell under direct internal reforming of liquid fuels. Chemical Engineering 2012;191:349–55.
- [37] Jamsak W, Assabumrungrat S, Douglas PL, Croiset E, Laosiripojana N, Suwanwarangkul R, et al. Performance assessment of bioethanol-fed solid oxide fuel cell system integrated with distillation column. ECS Transactions 2007;7(1):1475–82.
- [38] Baniasadi E, Dincer I. Energy and exergy analyses of a combined ammonia-fed solid oxide fuel cell system for vehicular applications. Hydrogen Energy 2011;36:11128–36.
- [39] Sanchez D, Chacartegui R, Torres M, Sanchez T. Stirling based fuel cell hybrid systems: an alternative for molten carbonate fuel cells. Journal of Power Sources 2008;192:84–93.
- [40] Calise F, d'Accadia MD, Vanoli L, von Spakovsky MR. Full load synthesis/design optimization of a hybrid SOFC-GT power plant. Energy 2007;32:446–58.
- [41] Fontell E, Kivisaari T, Christiansen N, Hansen JB, Pålsson J. Conceptual study of a 250 kW planar SOFC system for CHP application. Journal of Power Sources 2004;131(1–2):49–56.

Nomenclature

A_{ij} : matrix
 C: cost
 E_{FC} : electricity from fuel cell, V
 E_{Nernst} : Nernst ideal reversible voltage, V
 F: Faradays constant, C/mol
 f_{loss} : loss factor, –
 g^0 : standard Gibbs free energy, J/mol
 g_f : Gibbs free energy, J/mol
 h : enthalpy, J/kg
 h_f : enthalpy of formation, J/mol
 i_{as} : anode limiting current, mA/cm²
 i_d : current density, mA/cm²
 $\dot{n}_{H_2}$: molar reaction rate of H₂, mol/s
 P: power, W
 p: pressure, bar
 p_{H_2} : partial pressures for H₂, bar
 p_{H_2O} : partial pressures for H₂O, bar
 Q: heat, J/s
 T: temperature, K
 t: electrolyte thickness, m
 R: universal gas constant, J/kmol K
 U_F : universal gas constant, J/kmol K
 V: volume, m³
 V_{an} : anode porosity
 W: power consumption, W
 X_{H_2} : mass reaction rate of H₂
 y: molar fraction

Greek symbols

ΔE_{act} : activation polarization, V
 ΔE_{conc} : concentration polarization, V
 ΔE_{offset} : offset polarization, V
 ΔE_{ohm} : ohmic polarization, V
 ΔT : temperature difference, K
 ϵ : effectiveness
 ϵ_h : heater effectiveness
 ϵ_r : regenerator effectiveness
 η : efficiency
 η_{is} : isentropic efficiency
 η_m : mechanical efficiency
 η_{rev} : reversible efficiency
 η_v : voltage efficiency
 σ : conductivity, W/mK
 τ_{an} : anode tortuosity, m
 v: specific volume, m³/kg
 ξ : compression ratio
 ζ : temperature ratio

Subscripts

an: anode
 ca: cathode

cw: cold water
el: electrolyte
hw: hot water
ref: reference
s: entropy

Abbreviations

AP: anode preheater
ASR: adiabatic steam reformer
CHP: combined heat and power

CP: cathode preheater
CPO: catalytic partial oxidation
DME: di-methyl ether
DNA: dynamic network analysis
GT: gas turbine
HRS: heat recover steam generator
LHV: lower heating value
NG: natural gas
SOF: solid oxide fuel cell
ST: steam turbine